



# Production of fuel-cell grade H<sub>2</sub> by sorption enhanced steam reforming of acetic acid as a model compound of biomass-derived bio-oil



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## ARTICLE INFO

### Article history:

Received 21 September 2015

Received in revised form

17 November 2015

Accepted 18 November 2015

Available online 21 November 2015

### Keywords:

Hydrogen

Fuel cell

Bio-oil

Sorption enhanced steam reforming (SESR)

Pd/Ni–Co catalyst

## ABSTRACT

Fuel-cell grade H<sub>2</sub> has been produced by the sorption enhanced steam reforming (SESR) of acetic acid, a model compound of the bio-oil obtained from the fast pyrolysis of biomass. A Pd/Ni–Co catalyst derived from a hydrotalcite-like material (HT) with dolomite as CO<sub>2</sub> sorbent was used in the process. A fixed-bed reactor with three temperature zones was employed to favor the catalytic steam reforming reaction in the high-temperature segment, the SESR reaction in the intermediate-temperature part, as well as the water-gas shift (WGS) and CO<sub>2</sub> capture reactions in the low-temperature segment. Different conditions of pressure, temperature, steam/C molar ratio and weight hourly space velocity (WHSV) in the feed were evaluated. Higher steam/C molar ratios and lower WHSV values facilitated the production of H<sub>2</sub> and reduced the concentrations of CH<sub>4</sub>, CO and CO<sub>2</sub> in the produced gas. A fuel-cell grade H<sub>2</sub> stream with a H<sub>2</sub> purity of 99.8 vol.% and H<sub>2</sub> yield of 86.7% was produced at atmospheric pressure, with a steam/C ratio of 3, a WHSV of 0.893 h<sup>−1</sup> and a temperature of 575 °C in the intermediate part of the reactor (675 °C in the upper segment and 425 °C in the bottom part). At high pressure conditions (15 atm) a maximum H<sub>2</sub> concentration of 98.31 vol.% with a H<sub>2</sub> yield of 79.81% was obtained at 725 °C in the intermediate segment of the reactor (825 °C in the upper segment and 575 °C in the bottom part). Under these conditions an effluent stream with a CO concentration below 10 ppm (detection limit) was obtained at both low and high pressure, making it suitable for direct use in fuel cell applications.

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## 1. Introduction

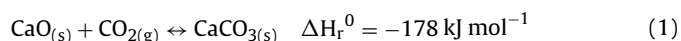
The worldwide demand for hydrogen for both chemical and energy uses is expected to increase in the medium-to-long term. Hydrogen is a clean fuel and energy carrier that, coupled with a fuel cell, has the potential to provide clean electricity due to the fact that both the fuel efficiency and level of CO<sub>2</sub> emissions of fuel cell-based systems are greatly improved compared to conventional power sources [1]. The most widespread and established technology for hydrogen production on a commercial scale nowadays is still the steam methane reforming (SMR) of natural gas. This reaction is performed at high temperature and high pressure, is strongly

endothermic and produces CO<sub>2</sub>. The process incorporates a high temperature catalytic SMR reactor (800–900 °C; 15–30 bar), one or two catalytic water-gas shift (WGS) reactors (200–400 °C) and a pressure swing adsorption (PSA) unit, since the product gas is usually fed to a PSA unit in order to obtain a high-purity (>99.9 vol.%) stream of H<sub>2</sub>. Hence, besides the consumption of fossil fuels, the overall process entails high capital costs [2]. This explains the increasing interest in the development of more energy-efficient and cost-competitive technologies for the production of hydrogen from renewable sources, such as biomass.

To overcome the disadvantages of SMR, the sorption enhanced steam reforming (SESR) process has received a great deal of attention in recent years. During the SESR process the steam reforming (SR), WGS and CO<sub>2</sub> capture reactions can be conducted simultaneously in one single reactor in conditions of moderate temperature and pressure. The CO<sub>2</sub> is removed in-situ by a sorbent as it is

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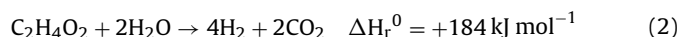
formed and hence the production of  $H_2$  is favored due to the shift of the reaction equilibrium towards hydrogen production. Different configurations of fixed or fluidized bed reactors have been proposed for carrying out the process. Basically the reactor is packed with a mixture of a reforming catalyst and a  $CO_2$  chemisorbent, while the feed gas (fuel + steam) is passed through the bed [3]. The  $CO_2$  by-product is removed from the reaction zone by the sorbent and a stream of pure  $H_2$  is produced at feed gas pressure. The sorbent needs to be periodically regenerated to allow a cyclic operation. The most commonly used sorbents are CaO-based, since they can repeatedly capture  $CO_2$  via consecutive CaO-carbonation and  $CaCO_3$ -calcination cycles as indicated by Eq. (1):



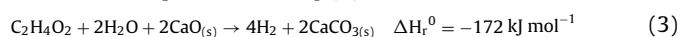
Natural CaO-based sorbents, such as limestone and dolomite, are able to react with  $CO_2$  at low  $CO_2$  partial pressures and moderate temperature, and show fast kinetics and good adsorption capacities. Although these materials usually suffer from decay in  $CO_2$  capture capacity after several carbonation/calcination cycles [4], they are preferably used in SESR studies due to their wide availability and very low cost [5–8].

Acetic acid is the fuel selected for the study of the SESR process in the present work, as it is one of the most representative constituents of the water soluble fraction of bio-oils produced from the fast pyrolysis of biomass, and hence is frequently used as a model compound of bio-oil in literature [9–13]. Since the bio-oil is derived from biomass resources, the catalytic steam reforming of bio-oil constitutes a renewable energy conversion method for producing hydrogen. Fast pyrolysis involves lower temperatures and shorter reaction times than gasification and produces mostly a liquid fraction. Bio-oil could be used for indirect cofiring in conventional power plants, direct decentralized heating purposes, and potentially as a high energy-density intermediate material suitable for long-distance transportation to destinations where it can be finally converted into chemicals and/or fuels. The use of bio-oil in the SESR process is advantageous because it has a higher energy density and it is easier to transport than raw biomass.

The overall reaction for the conversion of acetic acid to hydrogen, which is a combination of steam reforming and WGS reactions, is expressed by Eq. (2) as follows:



According to this reaction, 4 mol of hydrogen gas can theoretically be produced from the steam reforming of 1 mol of acetic acid. However, due to thermodynamic equilibrium limitations and complex reaction pathways, side reactions such as thermal decomposition or methanation usually occur, leading to the formation of intermediates and coke. Thus, the main products of the reforming process include  $H_2$ ,  $CO_2$ ,  $CH_4$  and CO. In the sorption enhanced  $H_2$  production process, the sorbent removes the  $CO_2$  from the gas phase as it is formed, promoting the direct WGS and reforming reactions due to an equilibrium shift effect. As a result, the CO and  $CH_4$  contents are significantly reduced while the  $H_2$  concentration and conversion are increased. Production of hydrogen with low levels of CO,  $CO_2$  and  $CH_4$  is critical for the generation of energy by means of hydrogen fuel cells [14]. The resulting reaction for the SESR process of acetic acid is represented in Eq. (3):



It can be seen that the SESR process with CaO as  $CO_2$  sorbent is exothermic and no additional energy is required in the reforming stage.

However, a very active catalyst must be used to obtain a high yield of very pure hydrogen from SESR for fuel-cell applications, since the temperatures used in this process are relatively low (400–650 °C). Furthermore, the catalyst must be stable, since it

goes through multiple oxidation/reduction cycles, corresponding to the calcination/carbonation stages [15]. In the present work, a Pd/Ni–Co catalyst has been synthesized and employed, since in a previous study it was demonstrated to be a promising material for use in the SESR process [16]. It shows a good reforming performance and allows the cycling operation of the sorption enhanced reforming process. Additionally, the Pd/Ni–Co catalyst makes the catalyst reduction step between the air-regeneration and reforming stages unnecessary due to the ability of Pd to promote the rapid production of  $H_2$  and hence facilitate an early reduction of the metal oxide phases in the catalyst.

The SESR process has been shown to improve the hydrogen production compared to conventional SR and satisfactory results have been reported in the literature after the SESR of methane [17–21]. Values of  $H_2$  purity as high as 94 vol.% have been obtained at atmospheric pressure, 650 °C and at a steam/C ratio of 3.2 using a NiO/NiAl<sub>2</sub>O<sub>4</sub> catalyst and a CaO-based synthetic sorbent, with a CO concentration as low as 2 vol.% [15]. Likewise, a  $H_2$  purity value of 95 vol.% has been reported at 15 atm, 725 °C and at steam/C ratio of 4 using a NiO/Al<sub>2</sub>O<sub>3</sub> catalyst and CaO as sorbent [22]. The benefits of the SESR of biomass compounds compared to SR have also been demonstrated with pyrolysis oils obtained from bunches of palm fruit and pine [23] or waste cooking oil [24]. Other experimental results from the SESR of biomass-derived compounds have been reported in the literature using mixtures of different Ni–Co HT catalysts and calcined dolomite as  $CO_2$  acceptor. These studies were carried out at atmospheric pressure, temperatures between 550 and 650 °C, and at steam/C molar ratios of 3–9. Thus, high-purity  $H_2$  (96.1–99.9 vol.%) with a low CO content (0.012–1.4 vol.%) was obtained from the SESR of bio-syngas from gasified biomass [25], bio-ethanol [26], lignocellulosic biomass [27], sugars from hydrolyzed biomass [28], pure glycerol [2,29] and crude glycerol, a by-product of biodiesel production [30]. However, the level of CO in the hydrogen stream obtained in these studies is too high for direct use in fuel cells. To produce very pure hydrogen directly from biomass in a single step is still a formidable challenge.

Low temperature proton exchange membrane (PEM) fuel cells require very low CO levels (<10 ppm) in a  $H_2$ -rich gas stream in order to be able to suppress the poisoning effect of CO on the PEM fuel cell catalysts [31,32]. The production of very pure hydrogen is therefore an important step for overcoming the CO poisoning problem. The WGS reaction is usually used in fuel processing for fuel cell applications with the purpose of reducing the level of CO. It is often followed by a preferential oxidation (PrOx) step for a final clean-up, where most of the residual CO is oxidized to  $CO_2$ . SESR could be an appropriate one-step high-purity hydrogen production process, which would greatly simplify the procedure, increase energy efficiency and reduce costs [33].

The objective of the present paper is to produce fuel cell-grade hydrogen in a single process, which has not been previously reported in the literature. To achieve this objective, the SESR of acetic acid, as a bio-oil model compound, using a Pd/Ni–Co catalyst and dolomite as  $CO_2$  sorbent was performed in a fixed bed reactor. Three zones of descending temperature were employed along the length of the reactor in order to be able to produce a high-purity  $H_2$  stream suitable for direct use in low-temperature PEM fuel cells. Thus, the proposed process involves favoring the catalytic reforming reaction in the high-temperature upper segment, the SESR reaction in the intermediate-temperature part, and the WGS and  $CO_2$  capture reactions in the low-temperature bottom segment. The influence of the operating parameters (pressure, temperature, steam/C molar ratio and space velocity) on the SESR of acetic acid was studied. The experimental results were compared with the equilibrium predictions obtained by means of the Aspen Plus process simulator under identical conditions to the experimental ones.

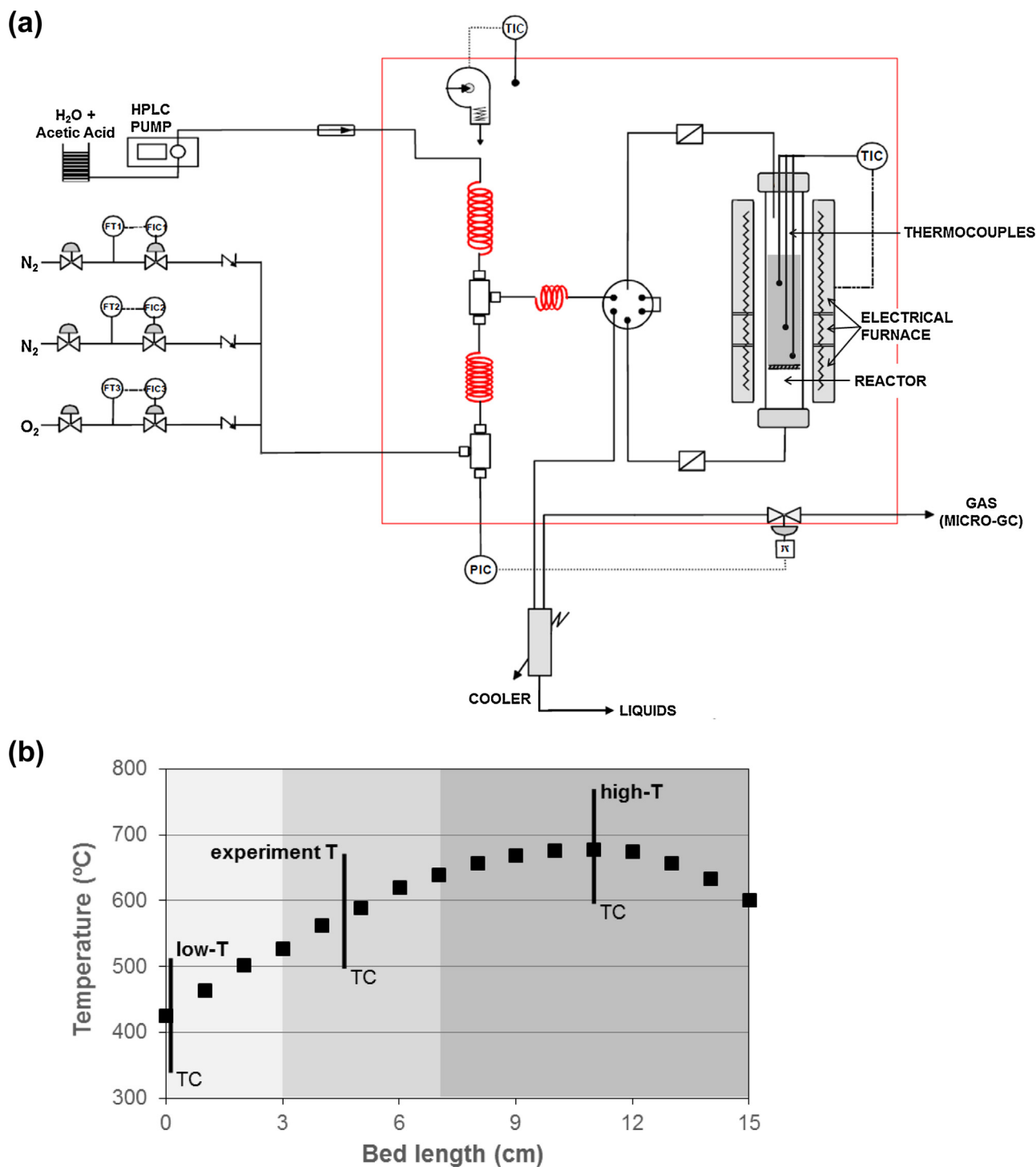


Fig. 1. (a) Schematic flow diagram of the experimental setup used for the SESR of acetic acid; (b) Temperature profile in the reactor (TC: thermocouple).

## 2. Experimental

### 2.1. Fuel

Acetic acid was selected as an oxygenated model compound of the organic acids present in the aqueous phase of bio-oils produced by the fast pyrolysis of biomass. Glacial acetic acid was supplied by PANREAC (100% purity). Aqueous solutions of acetic acid were prepared using water-to-acetic acid molar ratios of 2, 4 and 6 (steam/C molar ratios of 1, 2 and 3).

### 2.2. CO<sub>2</sub> sorbent

Arctic dolomite, used as a precursor of CaO for the capture of CO<sub>2</sub>, was supplied by Frønes Miljøkalk AS, Norway. It has a purity of 98.5 wt.% CaMg(CO<sub>3</sub>)<sub>2</sub> and does not contain any sulfur according to X-ray fluorescence analysis. The dolomite sample was calcined in an air flow (200 mL min<sup>-1</sup>) at 770 °C for 4 h prior to its use as CO<sub>2</sub> sorbent. Its initial maximum CO<sub>2</sub> capture capacity was estimated as being 0.46 g CO<sub>2</sub>/g sorbent. The dolomite was ground and sieved to reach a particle size of 125–250 μm. The BET surface area and pore

volume of the calcined dolomite are  $11.0 \text{ m}^2 \text{ g}^{-1}$  and  $0.16 \text{ cm}^3 \text{ g}^{-1}$ , respectively, while its average pore size is  $59 \text{ nm}$  [16].

### 2.3. Catalyst preparation

The 1%Pd/20%Ni–20%Co HT catalyst (Pd/Ni–Co HT) was prepared by the incipient wetness impregnation method using a 20%Ni–20%Co hydrotalcite-like material (Ni–Co HT) as precursor. The Ni–Co HT precursor was prepared by co-precipitation according to the method reported by He et al. [34] and it was impregnated with a 1% (w/w) load of Pd. The Pd solution was prepared by dissolving  $\text{PdCl}_2$  into two equivalents of HCl and diluting them in ethanol to the desired concentration. The sample was dried and calcined in an air flow at  $500^\circ\text{C}$  for 1 h in a muffle oven. The calcined catalyst was pelletized, ground and sieved to the desired particle size ( $125\text{--}250 \mu\text{m}$ ). It was then reduced at  $670^\circ\text{C}$  for 10 h in a mixed flow ( $100 \text{ mL min}^{-1}$ ) of 50 vol.% of  $\text{H}_2$  (balance  $\text{N}_2$ ). Finally, the catalyst was passivated by flowing 0.5 vol.% of  $\text{O}_2$  in a  $\text{N}_2$  stream for 1.5 h and then stored. A detailed description of the Pd/Ni–Co HT catalyst preparation procedure and characterization has been reported elsewhere [16,27]. Briefly, the Pd/Ni–Co HT catalyst has a BET surface area of  $144 \text{ m}^2 \text{ g}^{-1}$ . Its average pore size of  $12 \text{ nm}$  suggests that the catalyst is a mesoporous material. The metal dispersion estimated by chemisorption of  $\text{H}_2$  on the reduced catalyst is 7.8%, while the particle size and the metal surface area are  $13 \text{ nm}$  and  $21 \text{ m}^2_{\text{metal}} \text{ g}_{\text{catalyst}}^{-1}$ , respectively.

### 2.4. Sorption enhanced steam reforming (SESR) experiments

A schematic diagram of the experimental setup used for the catalytic SESR experiments with acetic acid is shown in Fig. 1a. It consists of a purpose-built downdraft fixed bed stainless steel reactor (with an internal diameter of  $13 \text{ mm}$  and a height of  $305 \text{ mm}$ ). The reactor is located inside a tubular electric furnace equipped with three temperature zones: an upper segment at the highest temperature, an intermediate part at the intermediate temperature and a lower part at the lowest temperature. The reaction temperatures are measured by K-type thermocouples which are inserted into the catalyst/sorbent bed. The pressure is measured by a pressure transducer and automatically controlled by a micro-valve. The gas flow rates are controlled by Bronkhorst® mass flow controllers, while the aqueous solution of acetic acid is fed in by means of a Gilson® high-performance liquid chromatography (HPLC) pump.

The reactor was loaded with a  $18 \text{ g}$  mixture of calcined dolomite (as  $\text{CO}_2$  sorbent) and Pd/Ni–Co catalyst, at a sorbent-to-catalyst ratio of  $8 \text{ g/g}$ . SESR experiments were carried out at 1 and 15 atm and isothermally at temperatures between  $525$  and  $775^\circ\text{C}$  in the intermediate segment of the reactor ( $40 \text{ mm}$ ). This intermediate temperature will be referred as ‘experiment temperature’ from now on, while the terms ‘high-temperature’ and ‘low-temperature’ will be used to refer to the temperature in the upper ( $80 \text{ mm}$ ) and lower ( $30 \text{ mm}$ ) segments of the reactor, respectively. The low-temperature is  $150^\circ\text{C}$  lower than the experiment temperature, while the high-temperature is  $100^\circ\text{C}$  higher than the experiment temperature. The temperature profile in the reactor and the location of the thermocouples are shown in Fig. 1b.

Prior to each SESR experiment, the catalyst/sorbent mixture was subjected to a regeneration step at  $770^\circ\text{C}$  and at 1 atm (or  $925^\circ\text{C}$  and 15 atm in the case of the high-pressure experiments) in an air flow ( $200 \text{ mL min}^{-1}$ ) until the  $\text{CO}_2$  level fell to less than 0.1 vol.%. The regeneration temperatures were selected taking into consideration the thermodynamic limitations of the decarbonation reaction and the kinetics of the decarbonation of dolomite [26]. Therefore, a regenerated catalyst/sorbent mixture was used in all the SESR experiments in the present study. After regeneration,

the reactor was purged with  $\text{N}_2$  and cooled down to the desired experimental reaction temperature. Once the operating temperature was reached under a  $\text{N}_2$  atmosphere, the acetic acid aqueous solution, swept by a  $50 \text{ mL min}^{-1}$  flow of  $\text{N}_2$ , was evaporated and then pumped downdraft through the catalyst/sorbent bed at different space velocities. The SESR of acetic acid, characterized by an enhanced production of hydrogen due to the removal of  $\text{CO}_2$  in situ by the sorbent (prebreakthrough stage), proceeded until the  $\text{CO}_2$  sorbent became saturated and lost its capacity for the removal of  $\text{CO}_2$  (breakthrough stage). Afterwards,  $\text{CO}_2$  capture was negligible (postbreakthrough stage) and a conventional SR process was assumed to take over.

The effluent gas exiting the reactor was directed into a thermoelectric cooler ( $2\text{--}4^\circ\text{C}$ ) to condense excess steam, unreacted acetic acid and any other liquid that may have formed. The composition of the dry gas was analyzed on-line by a dual channel Varian® CP-4900 Micro GC equipped with a molecular sieve (Molsieve 5 Å) column, a HayeSep A column and a thermal conductivity detector (TCD). Helium was used as the carrier gas. The Micro GC was calibrated employing a standard gas mixture at periodical intervals and the detection limit of the equipment for gas analysis is 0.001 vol.%. The species detected were  $\text{H}_2$ ,  $\text{CH}_4$ ,  $\text{CO}$ , and  $\text{CO}_2$ . Their concentrations in the gas product were calculated on the basis of the dry composition of the gas effluent. The flow rates of the species generated during the experiment were calculated by means of a nitrogen balance, on the basis of the amount of nitrogen fed in and the composition of the nitrogen evolved.

The  $\text{H}_2$  yield,  $\text{H}_2$  selectivity and the concentration of the gas components were calculated from Eq. (4)–(6), respectively:

$$\text{H}_2\text{yield}(\%) = 100 \times \left( \frac{F_{\text{H}_2}}{4 \times F_{\text{acetic acid}}} \right) \quad (4)$$

$$\text{H}_2\text{selectivity}(\%) = 100 \times \left( \frac{2 \times y_{\text{H}_2}}{2 \times y_{\text{H}_2} + 4 \times y_{\text{CH}_4}} \right) \quad (5)$$

$$i(\text{vol.}\%) = 100 \times \left( \frac{y_i}{\sum y_i} \right) \quad (i = \text{H}_2, \text{CH}_4, \text{CO}, \text{CO}_2) \quad (6)$$

where  $F_{\text{H}_2}$  is the molar flow rate of the  $\text{H}_2$  produced ( $\text{mol min}^{-1}$ ),  $F_{\text{acetic acid}}$  is the molar flow rate of the acetic acid fed in ( $\text{mol min}^{-1}$ ), and  $y_i$  is the molar content ( $\text{N}_2$  free and on a dry basis) of each species  $i$  ( $\text{H}_2$ ,  $\text{CH}_4$ ,  $\text{CO}$  and  $\text{CO}_2$ ). The weight hourly space velocity (WHSV) is defined as the ratio of the mass flow rate of the inlet acetic acid to the mass of catalyst ( $\text{g}_{\text{acetic acid}} \text{ g}_{\text{catalyst}}^{-1} \text{ h}^{-1}$ ).

### 2.5. Thermodynamic equilibrium calculations

A thermodynamic analysis of the SESR process was conducted at the three temperatures employed in the reactor (high-temperature, experiment temperature and low-temperature) under each one of the experimental reaction conditions studied. Aspen Plus 7.2 software (Aspentech) was used for the calculations. Equilibrium compositions were estimated by minimizing the Gibbs free energy, since this non-stoichiometric approach offers greater flexibility when tackling complex problems where the reaction pathways are unclear. A stoichiometric approach would require a clearly defined reaction mechanism incorporating all chemical reactions and species involved. The non-stoichiometric chemical equilibrium calculation method is incorporated in Aspen Plus as a RGibbs unit operation, which was chosen as the reaction system. RGibbs reactor block uses the criteria of minimization of the Gibbs free energy for all species in the system under phase and chemical equilibria without specification of the possible reactions. The Peng–Robinson property method was used to predict the thermodynamic behavior of the system. This calculation requires identification of the possible products. According to the results obtained for the prediction

of the equilibrium under sorption enhanced and steam reforming conditions, the species produced in concentrations higher than  $10^{-4}$  mol% were  $H_2$ ,  $CO$ ,  $CO_2$ ,  $CH_4$ ,  $H_2O$ ,  $CaO$  and  $CaCO_3$ .  $C_2H_4$  and  $C_2H_6$  were also included in the product pool, but their concentrations in the equilibrium stream were either null or not high enough to be considered significant products. The product mole fractions were calculated on a dry basis to compare with experimental results.

### 3. Results and discussion

#### 3.1. Effect of the temperature and pressure

##### 3.1.1. Thermodynamic equilibrium results

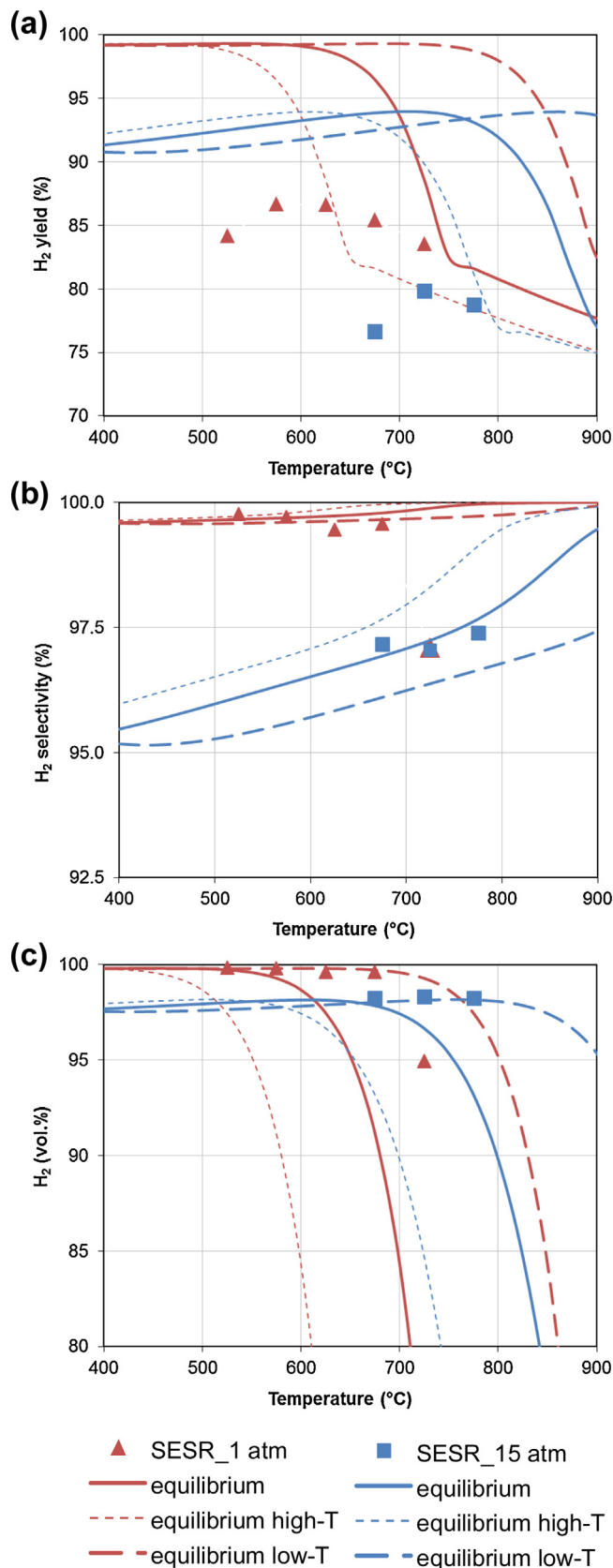
The thermodynamic equilibrium of the SESR process was analyzed at 1 and 15 atm and over a temperature range of 400–900 °C. Figs. 2–5 show the equilibrium values by means of lines corresponding to three temperatures: high-temperature corresponding to the upper segment of the reactor ('equilibrium high-T' lines), experiment temperature corresponding to the medium part of the reactor ('equilibrium' lines) and low-temperature corresponding to the lower part of the reactor ('equilibrium low-T' lines). The experiment temperature is the one shown on the X-axis. The points represent the experimental results.

In the case of the SESR process (Figs. 2 and 3), the experimental  $H_2$  concentration values (Fig. 2c) are higher, whereas the  $CO$  (Fig. 3a) and  $CO_2$  (Fig. 3b) concentrations are much lower, than those of the equilibrium corresponding to both the high-temperature and the experiment temperature. However, they all follow the lines of the equilibrium corresponding to the low-temperature quite closely between 525 and 675 °C in the case of the experiments at 1 atm and between 675 and 775 °C for the experiments at 15 atm. At atmospheric pressure and 725 °C, the  $H_2$  concentration value is below, while the  $CO$  and  $CO_2$  concentration values are above, the 'equilibrium low-T' curve.

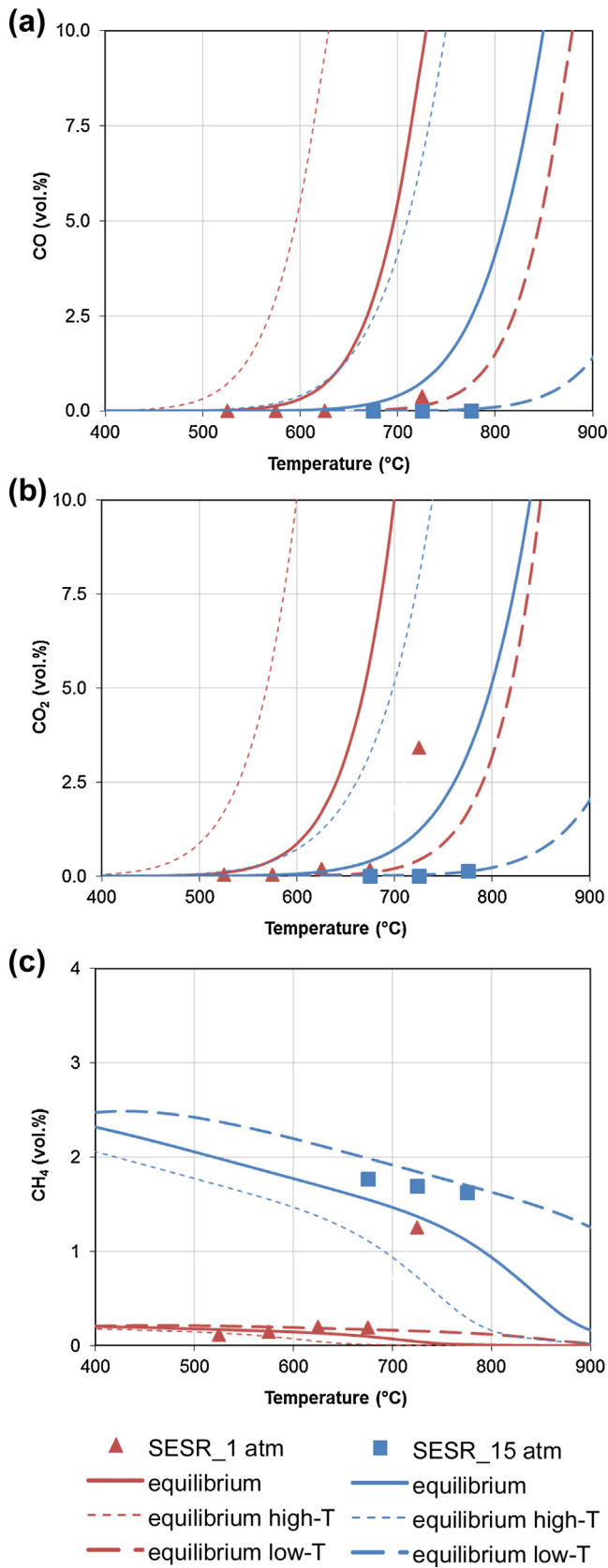
At atmospheric pressure the equilibrium values for  $CH_4$  (Fig. 3c) in the temperature range studied are low and very similar for all the temperatures studied, although they decrease slightly as the temperature increases. The experimental  $CH_4$  concentrations increase slightly between 525 and 675 °C at 1 atm, but they are all between the limits of the high-T and low-T equilibrium curves, as also occurs for the experimental  $H_2$  selectivity values (Fig. 2b). At atmospheric pressure and 725 °C, the  $CH_4$  concentration is above, while the  $H_2$  selectivity is below, the equilibrium curves. At high-pressure, the  $CH_4$  concentrations and the  $H_2$  selectivity values are between the equilibrium values corresponding to the high-temperature and low-temperature.

Finally, the experimental  $H_2$  yield values (Fig. 2a) are below the 'equilibrium high-T' curve between 525 and 625 °C at atmospheric pressure and between 675 and 775 °C at 15 atm, although this equilibrium is almost reached at the highest temperature. The values for the SESR experiments at atmospheric pressure and between 675 and 725 °C are between the 'equilibrium high-T' curve and that of the equilibrium corresponding to the experiment temperature. Therefore, the  $H_2$  yield values are far away from those corresponding to the 'equilibrium low-T' curve for all the experimental conditions analyzed.

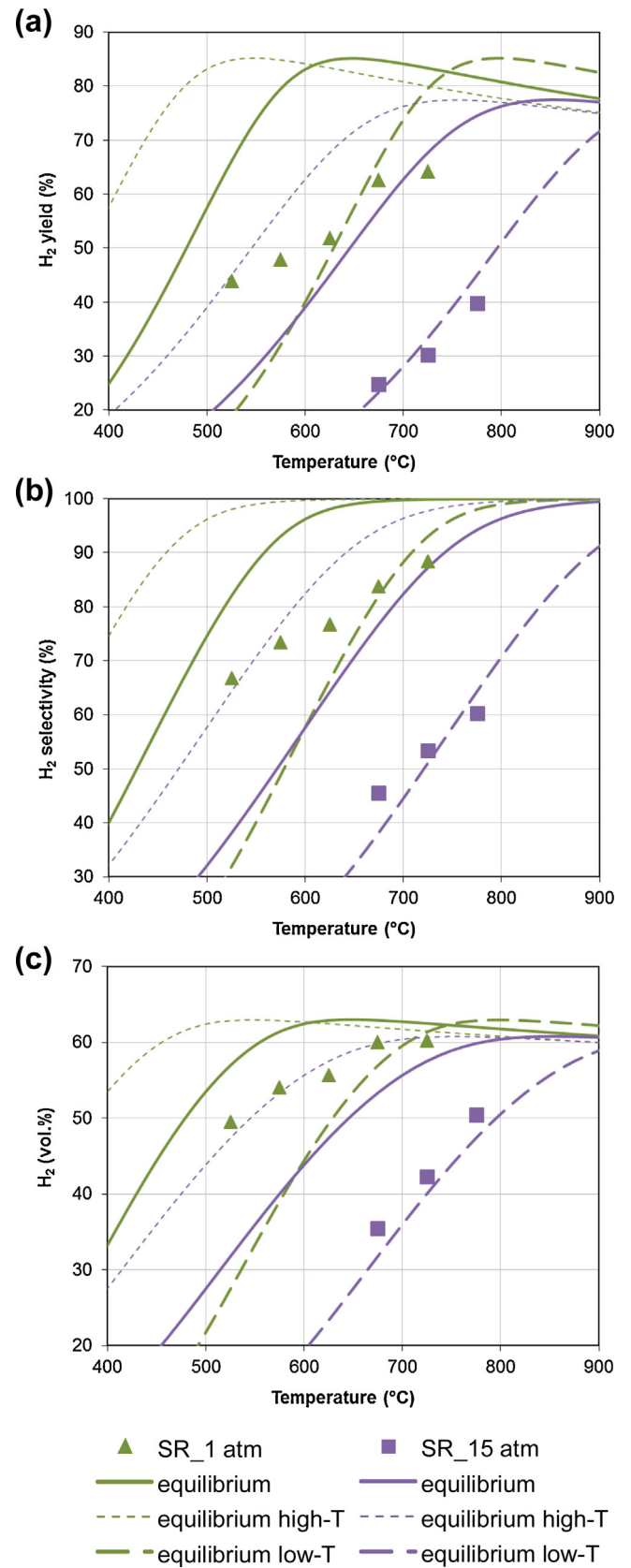
These results indicate that when the gas effluent crosses the bed in the lower part of the reactor (at the lowest temperature),  $CO_2$  capture increases and  $CO$  is converted by WGS, and so  $CO_2$  and  $CO$  concentrations decrease to values corresponding to the equilibrium at the lowest temperature ('equilibrium low-T' curve). Both the WGS reaction and the steam reforming reaction are enhanced by the capture of  $CO_2$ , which leads to an increase in  $H_2$  concentra-



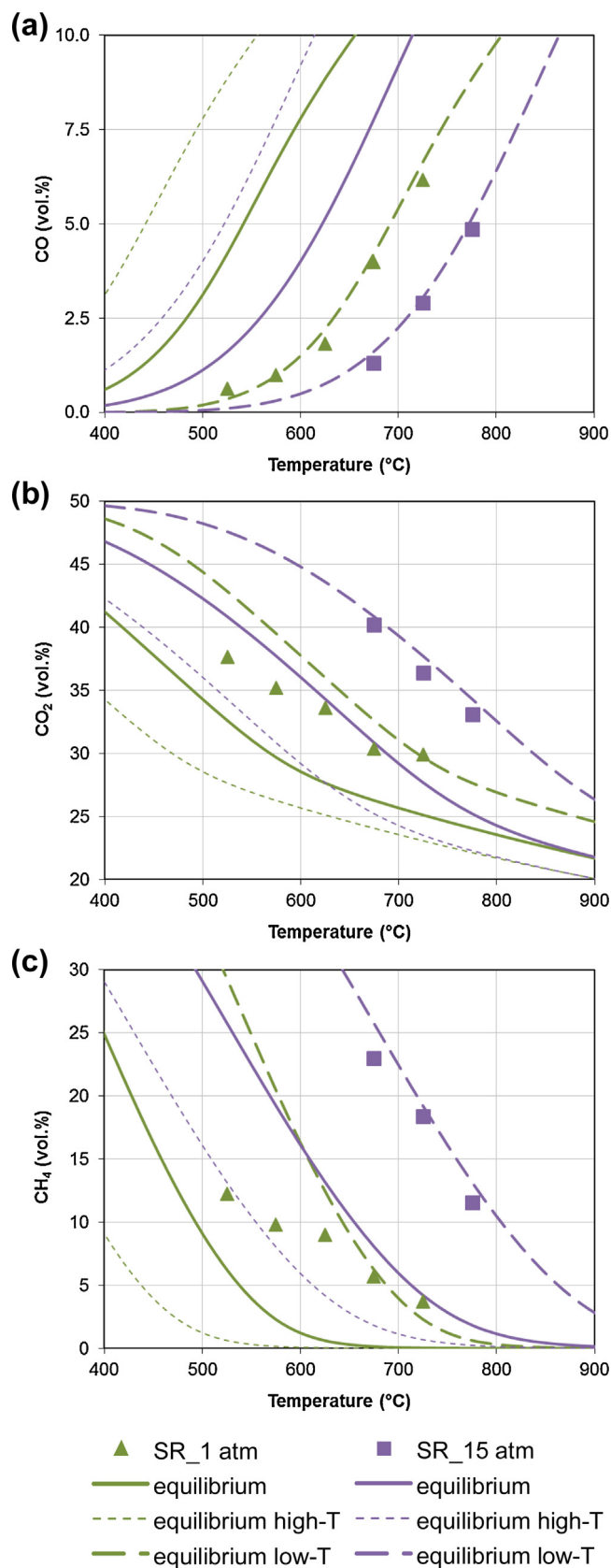
**Fig. 2.** Effect of the temperature and pressure on the  $H_2$  yield (a),  $H_2$  selectivity (b) and  $H_2$  concentration (c) during the SESR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, WHSV =  $0.893\text{ h}^{-1}$ , sorbent/catalyst ratio = 8 g/g, Pd/Ni-Co HT catalyst and dolomite as sorbent.



**Fig. 3.** Effect of the temperature and pressure on the CO (a), CO<sub>2</sub> (b) and CH<sub>4</sub> (c) concentrations during the SESR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, WHSV = 0.893 h<sup>-1</sup>, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.



**Fig. 4.** Effect of the temperature and pressure on the H<sub>2</sub> yield (a), H<sub>2</sub> selectivity (b) and H<sub>2</sub> concentration (c) during the SR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, WHSV = 0.893 h<sup>-1</sup>, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.



**Fig. 5.** Effect of the temperature and pressure on the CO (a), CO<sub>2</sub> (b) and CH<sub>4</sub> (c) concentrations during the SR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, WHSV = 0.893 h<sup>-1</sup>, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.

tion, helping to reach the equilibrium value corresponding to the low-temperature.

In general, similar trends regarding to the equilibrium were obtained in the case of the SR process (Figs. 4 and 5) than those for the SESR process. The experimental CO (Fig. 5a), CO<sub>2</sub> (Fig. 5b) and CH<sub>4</sub> (Fig. 5c) concentrations follow the ‘equilibrium low-T’ curve quite accurately, or they are close, for the experiments at both 1 and 15 atm. These results indicate that the lower temperature maintained during the experiments in the lower part of the reactor clearly favors the conversion of CO via the WGS reaction during the SR process. For the H<sub>2</sub> yield (Fig. 4a), H<sub>2</sub> selectivity (Fig. 4b) and H<sub>2</sub> concentration (Fig. 4c), their values are higher than those of the ‘equilibrium low-T’ curve at the lower temperatures studied, but slightly lower in the higher temperature range. Therefore, the SR results indicate that the SR reaction occurred in the upper part of the reactor where the temperature was high enough, but it is probable that the H<sub>2</sub> equilibrium concentrations corresponding to that temperature were not reached. This seems to be confirmed by the fact that the CH<sub>4</sub> concentrations are above the values corresponding to the equilibrium at the highest temperature. However, as the temperature was reduced in the lower part of the reactor, the WGS was favored and the consumption of CO promoted, which can be deduced from the experimental CO concentrations that are similar to those of the low-temperature equilibrium. This implies an increase in the H<sub>2</sub> concentration, H<sub>2</sub> selectivity and H<sub>2</sub> yield, which are higher than those of the low-T equilibrium.

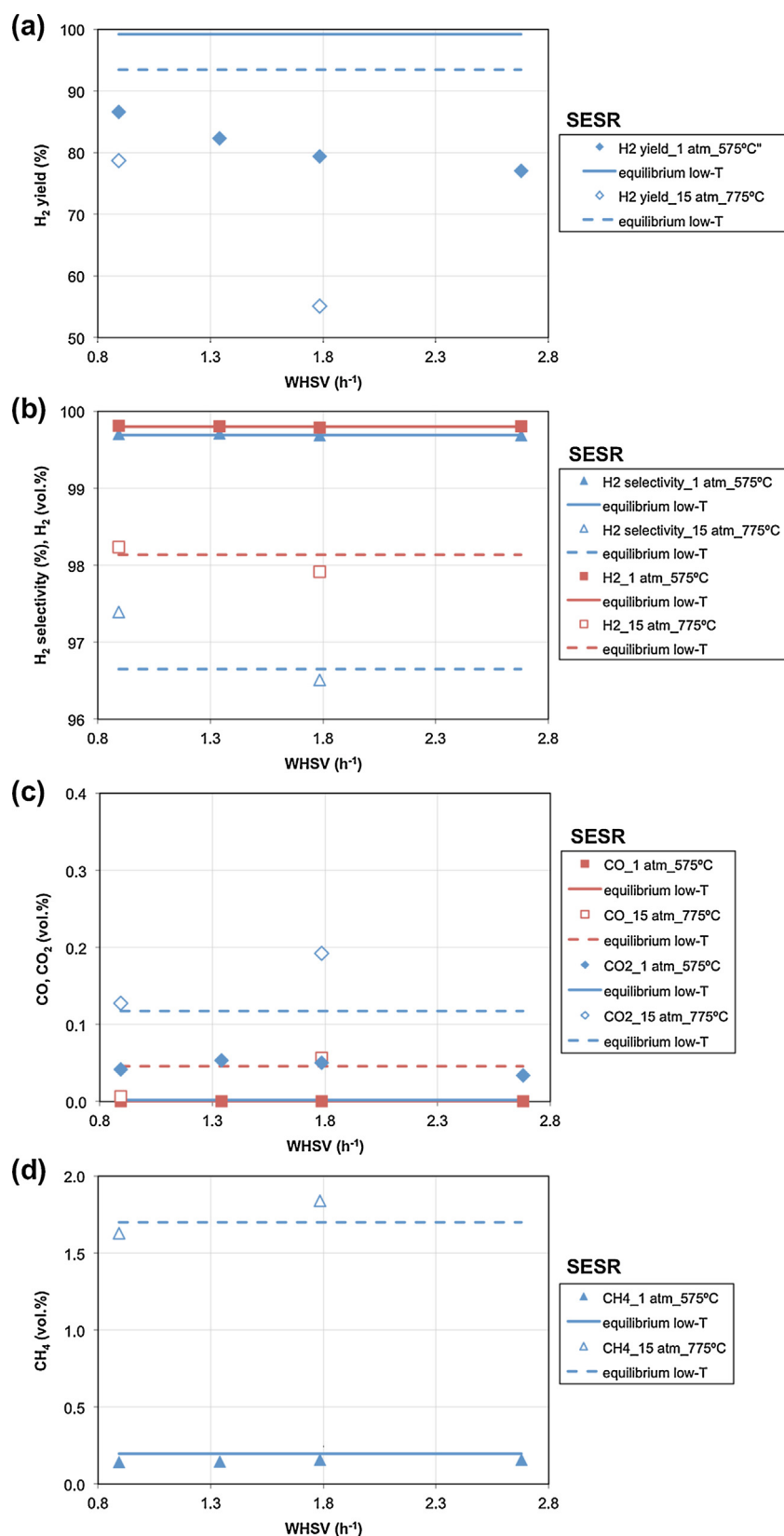
Therefore, for both SESR and SR the comparison between the equilibrium and experimental results suggests that the effluent composition is determined by the temperature at the reactor outlet. Thermodynamic equilibrium results show that the low temperature at the lower segment of the reactor favors the WGS reaction and hence the hydrogen purity, which demonstrates the advantages of the purposed strategy for the production of highly pure hydrogen by using three-temperature zones in the reactor.

### 3.1.2. Experimental results of the SESR process

Figs. 2 and 3 show the experimental results as a function of the experiment temperature during the SESR process. The H<sub>2</sub> yield (Fig. 2a) increases with the temperature up to 86.7% (at 575 °C and atmospheric pressure) and 79.8% (at 725 °C and 15 atm), evidencing a slight decrease at higher temperatures. The H<sub>2</sub> selectivity (Fig. 2b) shows very high values (>99.6%) at temperatures between 525 and 675 °C at atmospheric pressure and it decreases at 725 °C (97.1%). However, at 15 atm the H<sub>2</sub> selectivity values are very similar for the range of temperatures studied (around 97.2%).

At atmospheric pressure the H<sub>2</sub> concentration (Fig. 2c) reaches values as high as 99.8 vol.% at 525 and 575 °C, and 99.6 vol.% at 625 and 675 °C. Further increase in the experiment temperature up to 725 °C leads to a significant reduction in the H<sub>2</sub> purity (95.0 vol.%). However, similar concentrations of H<sub>2</sub> are obtained when SESR is carried out at 15 atm (98.3 vol.%) irrespective of the temperature. The different trends with temperature for the H<sub>2</sub> yield and H<sub>2</sub> concentration indicate that a compromise between these parameters must be accepted in order to establish an optimum temperature for the SESR process under the conditions studied.

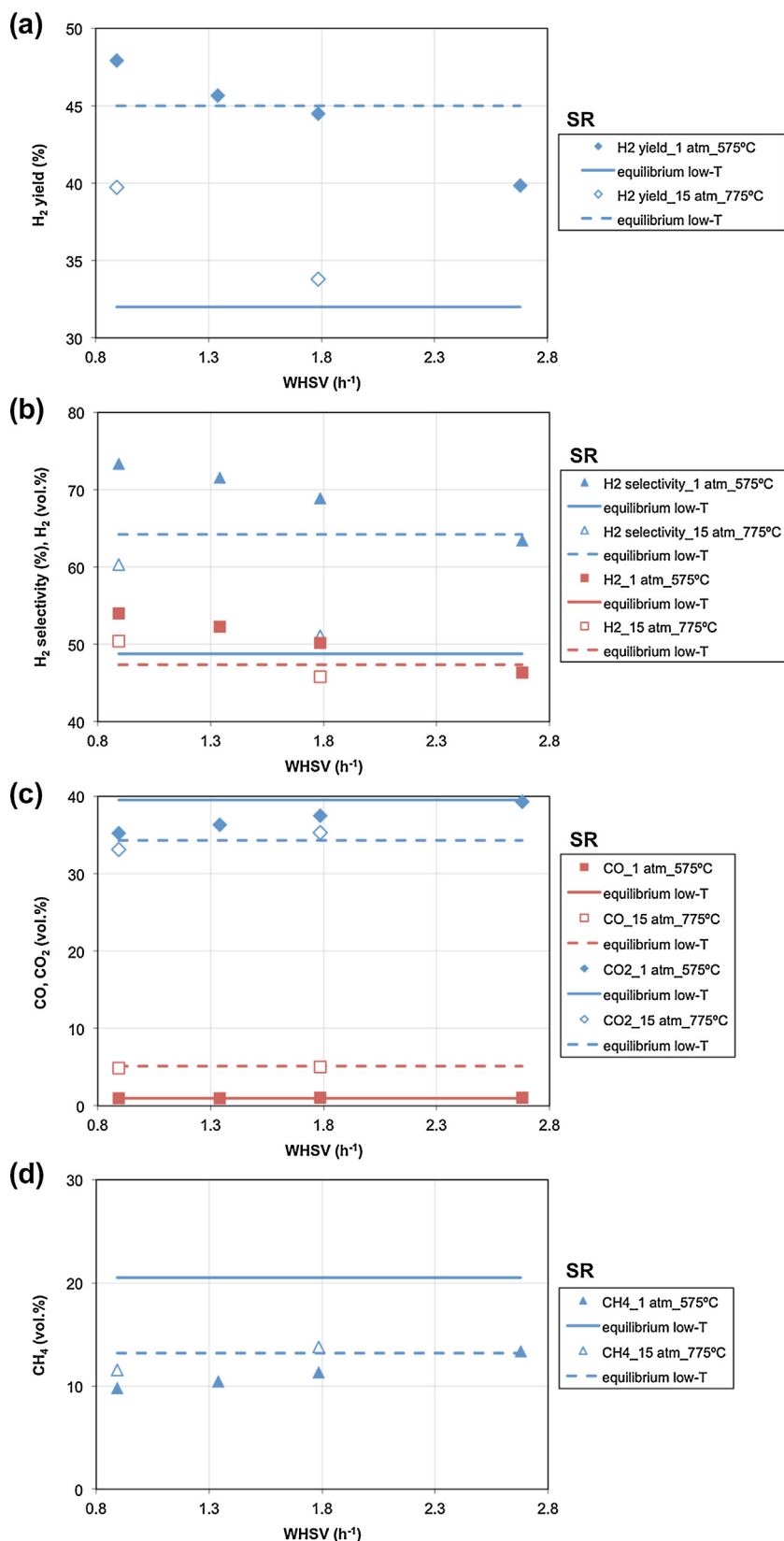
The CO concentration (Fig. 3a) is below the detection limit when the temperature is between 525 and 625 °C at 1 atm, while a low concentration is observed at 675 °C (0.02 vol.%). However, a significant value of the CO concentration (0.38 vol.%) is detected at 725 °C. At 15 atm, CO concentration is below the detection limit when the reaction occurs at 675 and 725 °C, while only a low CO concentration (0.01 vol.%) is apparent at 775 °C at high pressure. The favorable thermodynamics of the WGS reaction (exothermic) at low temperature explains the lower CO concentrations found at the lowest temperatures studied.



**Fig. 6.** Effect of the weight hourly space velocity (WHSV) and pressure on the H<sub>2</sub> yield (a), H<sub>2</sub> selectivity and H<sub>2</sub> concentration (b), CO and CO<sub>2</sub> concentrations (c) and CH<sub>4</sub> concentration (d) during the SESR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.

At atmospheric pressure, the CO<sub>2</sub> concentration (Fig. 3b) is very low when SESR takes place at 525 and 575 °C (0.04 vol.%). It increases slightly when the temperature increases up to

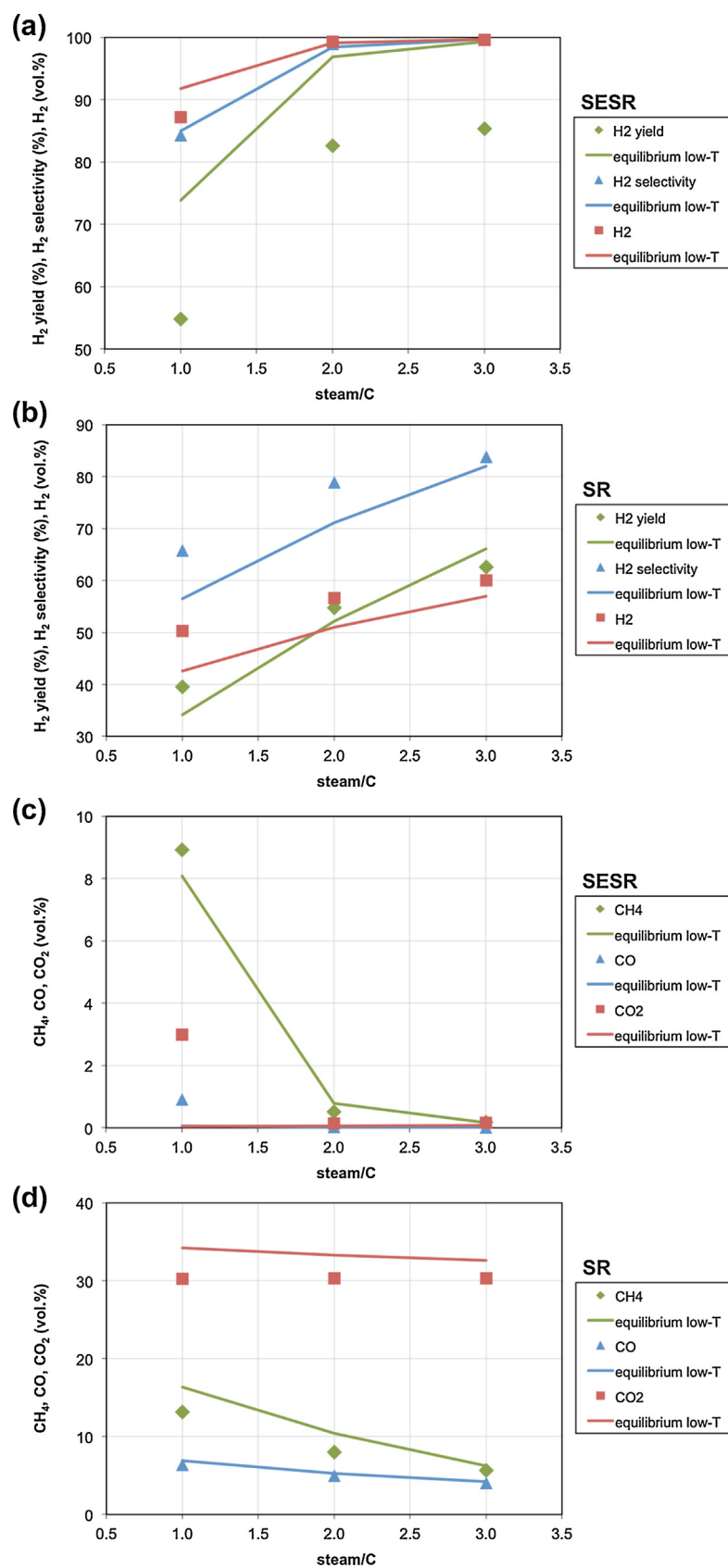
675 °C (0.18 vol.%), but it experiences a greater increase at 725 °C (3.4 vol.%). For SESR at 15 atm, the CO<sub>2</sub> concentration is below the detection limit at 675 and 725 °C, while a low concentration



**Fig. 7.** Effect of the weight hourly space velocity (WHSV) and pressure on the  $\text{H}_2$  yield (a),  $\text{H}_2$  selectivity and  $\text{H}_2$  concentration (b), CO and  $\text{CO}_2$  concentrations (c) and  $\text{CH}_4$  concentration (d) during the SR of acetic acid. Reaction conditions: steam/C = 3 mol/mol, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.

value is recorded at 775 °C (0.13 vol.%). In this case, the favorable thermodynamics of the carbonation reaction (exothermic) at low temperature explains why the  $\text{CO}_2$  concentration increases

with increasing temperature since high temperatures inhibit the removal of  $\text{CO}_2$  to some extent. This means that a very weak  $\text{CO}_2$  sorption occurs when the experiment temperature reaches 725 °C



**Fig. 8.** Effect of the steam/C molar ratio on the SESR (a and c) and SR (b and d) of acetic acid. Reaction conditions:  $P = 1$  atm,  $T = 675$  °C,  $WHSV = 0.893\text{--}1.563$  h<sup>-1</sup>, sorbent/catalyst ratio = 8 g/g, Pd/Ni–Co HT catalyst and dolomite as sorbent.

at atmospheric pressure, resulting in a poor sorption enhancement of the process. Since  $\text{CO}_2$  sorption leads to an enhancement in  $\text{H}_2$  production by shifting the equilibrium of the steam reforming and WGS reactions to the products side, lower values of  $\text{H}_2$  concentration and higher concentrations of  $\text{CO}$  are also found at the highest temperature studied.

The  $\text{CH}_4$  concentration (Fig. 3c) increases slightly (from 0.12 to 0.20 vol.%) as the temperature increases from 525 to 675 °C at 1 atm, whereas a much higher  $\text{CH}_4$  content is detected at 725 °C (1.25 vol.%). The low  $\text{CH}_4$  concentrations below 675 °C are indicative of the ability of the Pd/Co–Ni HT catalyst to suppress the methanation reaction due to the enhancement of the WGS reaction and/or to the successful catalysis of the methane steam reforming reaction. An approximately constant  $\text{CH}_4$  concentration is maintained when the SESR process occurs between 675–775 °C at 15 atm (1.7 vol.%). In general, these results are in good agreement with those obtained for the SESR of other biomass fuels, such as ethanol [26] or sorbitol and glucose [28].

In summary, the results indicate that an increase in the temperature above 675 °C during the SESR of acetic acid at atmospheric pressure significantly decreases the  $\text{H}_2$  yield,  $\text{H}_2$  selectivity and  $\text{H}_2$  concentration, while simultaneously increases the  $\text{CH}_4$ ,  $\text{CO}$  and  $\text{CO}_2$  concentrations. On the other hand, no significant differences are detected in the results when the SESR process of acetic acid is performed at 15 atm between 675 and 775 °C. However, if a fuel-cell grade stream of  $\text{H}_2$  (i.e.,  $\text{CO}$  concentration <10 ppm) is required, the SESR of acetic acid should be performed between 525 and 625 °C in the intermediate segment of the reactor at atmospheric pressure or between 675 and 725 °C at 15 atm.

Figs. 2 and 3 also show higher values for the  $\text{H}_2$  yield,  $\text{H}_2$  selectivity and  $\text{H}_2$  concentration, as well as lower values for the  $\text{CH}_4$  concentration, when the process is carried out at 675 °C and at atmospheric pressure compared to the high-pressure conditions (15 atm). Atmospheric pressure is thermodynamically favorable for producing a high concentration of  $\text{H}_2$ . An increase in the pressure does not favor the production of hydrogen, whereas it promotes the formation of methane by the methanation reaction. Since low pressure is also favorable for the decarbonation reaction of the solid  $\text{CO}_2$  acceptor, most sorption enhanced reforming studies have been performed at atmospheric pressure. However, the production of  $\text{H}_2$  at atmospheric pressure might lead to an energy penalty, as compressed  $\text{H}_2$  is required for its transportation and storage [35]. However, it should be noted that at 675 °C low concentrations of  $\text{CO}$  and  $\text{CO}_2$  are produced at atmospheric pressure, whereas these gases are below the detection limit in the effluent at 15 atm. When the experiment temperature is increased up to 725 °C, the  $\text{H}_2$  yield is only slightly higher at 1 atm, whereas the  $\text{H}_2$  concentration is much higher at 15 atm due to the very high concentrations of  $\text{CO}$  and  $\text{CO}_2$  present in the effluent at atmospheric pressure. The results indicate that at higher pressures, the temperature needs to be elevated considerably to achieve the same level of  $\text{H}_2$  concentration as that obtained at atmospheric pressure. The  $\text{CO}$  and  $\text{CO}_2$  concentrations are still below the detection limit when the process is carried out at 725 °C and 15 atm. As a result, at high pressures,  $\text{CH}_4$  is the dominant impurity in the production of hydrogen.

### 3.1.3. Experimental results of the SR process

The experimental results corresponding to the SR process are shown in Figs. 4 and 5. It can be observed that the  $\text{H}_2$  yield (Fig. 4a),  $\text{H}_2$  selectivity (Fig. 4b),  $\text{H}_2$  concentration (Fig. 4c) and  $\text{CO}$  concentration (Fig. 5a) increase as the experiment temperature increases. Also, their values are significantly higher when the reforming process is performed at atmospheric pressure compared to high-pressure conditions. As a result of the endothermic character of the reforming reaction, higher temperatures favor the conversion of fuel and hence lead to a higher concentration of  $\text{H}_2$ .

On the other hand, the  $\text{CO}_2$  (Fig. 5b) and  $\text{CH}_4$  (Fig. 5c) concentrations decrease with temperature, and their values are lower for the experiments performed at atmospheric pressure than those at 15 atm. The exothermic character of the WGS reaction leads to higher concentrations of  $\text{CO}$  and lower concentrations of  $\text{CO}_2$  at higher temperatures. Finally, the endothermic character of the reforming reaction of  $\text{CH}_4$  would explain the lower concentration of this component at higher temperatures.

In the case of the SR process, the difference between the results obtained at atmospheric pressure and 15 atm is much greater than that observed for the SESR process. Moreover, a clear worsening of the results when the process is carried out at higher pressure is evident at all the temperatures.

### 3.1.4. Optimum conditions for the SESR of acetic acid

In the SESR process, the highest experimental value of  $\text{H}_2$  concentration obtained is 99.84 vol.% at a temperature of 525 °C in the intermediate segment of the reactor. In this case, the  $\text{H}_2$  yield value obtained is 84.19% and the  $\text{CO}$  concentration is below the detection limit. The  $\text{H}_2$  yield can be increased up to 86.66% if the temperature in the intermediate segment of the reactor is increased up to 575 °C. At this temperature the  $\text{H}_2$  concentration (99.82 vol.%) does not experience any notable decrease, while the  $\text{CO}$  concentration is still below the detection limit.

On the other hand, if the SESR process is to be carried out at 15 atm, the most suitable temperature for the intermediate segment of the reactor would be 725 °C and, under such conditions, the  $\text{H}_2$  yield would be 79.81% and the  $\text{H}_2$  concentration 98.31 vol.%, whereas the  $\text{CO}$  concentration is below the detection limit.

The SESR of acetic acid was previously studied in a fluidized bed reactor at atmospheric pressure and at 575 °C, with a steam/C ratio of 3 and a WHSV of 0.893 h<sup>−1</sup> [36]. The results obtained were a  $\text{H}_2$  yield of 90.18%, a  $\text{H}_2$  selectivity of 99.69% and a  $\text{H}_2$  concentration of 99.29 vol.%. A  $\text{CO}$  content of 0.11 vol.% was also detected. Therefore, although a higher  $\text{H}_2$  yield than that of the present study was obtained, the  $\text{H}_2$  purity was lower and a fuel-cell grade  $\text{H}_2$  stream was not achieved due to the high  $\text{CO}$  concentration. This clearly indicates that an additional last stage of low temperature during the SESR process is effective to reduce the  $\text{CO}$  content in the effluent gas.

In summary, from the results obtained in the present work it can be deduced that at atmospheric pressure, a temperature of 575 °C in the intermediate segment of the reactor (675 °C in the upper segment of the reactor and 425 °C in the lower part), at a steam/C ratio of 3 with a WHSV of 0.893 h<sup>−1</sup>, can be considered as the optimum conditions for the SESR of acetic acid using the fixed-bed reactor configuration of the present study in order to be able to achieve a very high  $\text{H}_2$  production and to obtain a fuel-cell grade hydrogen stream.

### 3.2. Effect of weight hourly space velocity (WHSV)

The effect of the space velocity on the SESR and SR processes was evaluated at atmospheric pressure (at 575 °C) and under 15 atm (at 775 °C) conditions. WHSV values of 0.893, 1.339, 1.786 and 2.679 h<sup>−1</sup> were studied by changing the total inlet flow rate (5, 7.5, 10 and 15 g h<sup>−1</sup>, respectively), while maintaining a steam/C molar ratio of 3. Fig. 6 shows the results for the SESR process, whereas Fig. 7 shows the results for SR, as a function of the WHSV.

For the SESR, the  $\text{H}_2$  yield (Fig. 6a) decreases significantly with the space velocity in the experiments at both atmospheric and high-pressure. At atmospheric pressure, the  $\text{H}_2$  selectivity and  $\text{H}_2$  concentration (Fig. 6b), as well as the  $\text{CO}$ ,  $\text{CO}_2$  (Fig. 6c) and  $\text{CH}_4$  (Fig. 6d) concentrations are almost constant in the range of the space velocities studied. Furthermore, the  $\text{CO}$  concentration is below the detection limit under such conditions. However, when

the SESR process is carried out at 15 atm, the H<sub>2</sub> selectivity and H<sub>2</sub> concentration decrease with space velocity, while the CO, CO<sub>2</sub> and CH<sub>4</sub> concentrations increase. A lower fuel conversion might be expected with an increase in WHSV due to the shorter contact time. In the present work, a worse conversion is reflected in the decrease in the H<sub>2</sub> yield with the space velocity at both pressures. However, at atmospheric pressure the H<sub>2</sub> purity and the CO content are not affected by the increase in WHSV up to 2.679 h<sup>-1</sup>. Fig. 6 shows that the H<sub>2</sub> yield is affected much more by WHSV than H<sub>2</sub> selectivity and H<sub>2</sub> purity. Similar results for the hydrogen yield were found after the sorption enhanced steam reforming of glycerol [30]. This suggests that, together with a decrease in the conversion, the formation of coke may have occurred at higher space velocities.

In the case of the SR process, the H<sub>2</sub> yield (Fig. 7a), H<sub>2</sub> selectivity and H<sub>2</sub> concentration (Fig. 7b) decrease, while the CO, CO<sub>2</sub> (Fig. 7c) and CH<sub>4</sub> (Fig. 7d) concentrations increase, as the space velocity increases in the experiments at both 1 and 15 atm.

### 3.3. Effect of steam/C molar ratio

The effect of the steam/C molar ratio on the SESR and SR processes was evaluated at 675 °C and under atmospheric pressure conditions. Steam/C molar ratios of 1, 2 and 3 were studied by varying the inlet flow of acetic acid (3.13, 2.17 and 1.79 g h<sup>-1</sup>, respectively), which ensured that the total inlet flow rate of aqueous solution would remain the same (5 g h<sup>-1</sup>). The experimental results are shown in Fig. 8.

For both the SESR (Fig. 8a) and SR (Fig. 8b) processes, the H<sub>2</sub> yield, H<sub>2</sub> selectivity and H<sub>2</sub> concentration increase with the steam/C molar ratio, mainly when the steam/C molar ratio increases from 1 to 2. In the case of the SESR process, when the steam/C molar ratio increases from 2 to 3, the increase in H<sub>2</sub> selectivity (from 98.95 to 99.56%) and H<sub>2</sub> concentration (from 99.31 to 99.62 vol.%) is relatively small, due to thermodynamic limitations. However, a more significant increase is detected in the case of the H<sub>2</sub> yield (from 82.68 to 85.40%) as the steam/C molar ratio increases up to 3, which can be due to the fact that higher steam/C ratios favor the reaction kinetics, lowering the coke formation. In general, for SESR (Fig. 8c) and SR (Fig. 8d) the CH<sub>4</sub>, CO and CO<sub>2</sub> concentrations decrease as the steam/C molar ratio increases, but their decrease is again much higher when the steam/C molar ratio increases from 1 to 2. Low steam/C ratio values enhance methanation, whereas they cause a simultaneous decrease in methane steam reforming and WGS reactions [31]. Thus, the CH<sub>4</sub> concentration was reduced from 8.93 to 0.53 vol.% when the steam/C molar ratio increased from 1 to 2. Methane content is a very important factor for hydrogen selectivity in the SESR process. Likewise, at a low steam/C molar ratio, hydrogen purity is strongly influenced by the methane content. A lower hydrogen production at low values of steam/C ratio has been reported after the SESR of glycerol due to the formation of carbonaceous species on the catalyst that caused a decrease in its activity, so increasing the methane content at the expense of hydrogen production [29]. CO<sub>2</sub> and CO concentrations are almost constant when the steam/C molar ratio increases from 2 to 3 in the case of the SESR process, which indicates that the carbonation reaction is proceeding normally at the studied temperature at both steam/C ratio values.

An optimal steam/C molar ratio of 4.5 at atmospheric pressure, a temperature of 560 °C and a WHSV of 0.893 h<sup>-1</sup> have been previously reported in the literature [36] as the ideal conditions for the SESR of acetic acid for hydrogen production in a fluidized bed reactor. This steam/C molar ratio is higher than that used in the present work, and consequently a higher H<sub>2</sub> yield (92.0%) was obtained. However, a slightly lower H<sub>2</sub> purity was achieved (99.53 vol.%) with a higher CO content (0.06 vol.%) in the effluent stream, again show-

ing the advantage of adding an additional low-temperature stage to the reforming process.

Steam is often added beyond the stoichiometric limit in the reforming process in order to promote hydrogen production and to prevent coke formation over the surface of the catalyst. The effect of increasing the steam/C molar ratio is especially important for the H<sub>2</sub> yield (Fig. 8), which is significantly enhanced at higher values of steam/C. A higher amount of steam will increase the conversion of tars and intermediate compounds, and hence contribute to the avoidance of carbon deposits formation during the reforming process. In the present work, the experimental results indicate that catalytic activity is maintained when a steam/C molar ratio of 2 is used. H<sub>2</sub> purity can also be maintained at a reasonably high value under such conditions, with low CO and CO<sub>2</sub> concentrations. If a decrease in the H<sub>2</sub> yield were acceptable, a lower steam/C molar ratio could be used to reduce the heat requirements for steam generation and the associated energy penalty.

## 4. Conclusions

In order to produce a fuel cell-grade H<sub>2</sub> stream in a single step, the sorption enhanced steam reforming of acetic acid, a model compound of bio-oil obtained from biomass fast pyrolysis, was experimentally investigated. The experimental results demonstrate the possibility of obtaining high-purity H<sub>2</sub> after the SESR of acetic acid in a fixed bed reactor with three-temperature zones over a Pd/Ni–Co HT catalyst using calcined dolomite as CO<sub>2</sub> sorbent. The catalyst used proved to be highly active for the one-step production of fuel-cell grade hydrogen (<10 ppm CO) from the SESR of acetic acid. The optimum conditions for the production of hydrogen by the SESR of acetic acid were found to be 1 atm and 575 °C in the intermediate part of the reactor (675 °C in the upper segment and 425 °C in the lower part) at a steam/C molar ratio of 3 and a WHSV value of 0.893 h<sup>-1</sup>. Thus, a H<sub>2</sub> concentration of 99.82 vol.%, with a H<sub>2</sub> yield of 86.66% and a H<sub>2</sub> selectivity of 99.71%, were achieved. Low CH<sub>4</sub> and CO<sub>2</sub> concentrations (0.14% and 0.04 vol.%, respectively) and a negligible CO content (below detection limit) were also obtained. For high-pressure conditions (15 atm), a maximum H<sub>2</sub> concentration of 98.31 vol.%, with a H<sub>2</sub> yield of 79.81% and a H<sub>2</sub> selectivity value of 97.04%, can be obtained at 725 °C in the intermediate segment of the reactor (825 °C in the upper segment and 575 °C in the bottom part), together with a low CH<sub>4</sub> concentration (1.69 vol.%) and non-detectable levels of CO and CO<sub>2</sub>. Such a low level of CO in the H<sub>2</sub> stream might allow the direct use of the gas product in low-temperature proton exchange membrane fuel cells without the need for subsequent CO removal. From these results it can be concluded that an additional last stage of low temperature during the SESR process would produce a highly pure H<sub>2</sub> stream suitable for direct application in PEM fuel cells.

## Acknowledgements

This work was carried out with financial support from the Spanish MINECO (Project ENE2014-53515-P), co-financed by the European Regional Development Fund (ERDF) and the Principado de Asturias (PCTI 2013-2017, GRUPIN14-079). The authors thank Franeoss Miljøkalk A/S (Norway) for supplying Arctic dolomite.

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